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Modelling and control of nonlinear compressor unit in biohydrogen plant using multivariable model predictive control (MMPC)

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ABSTRACT

Biohydrogen plants consist of several units, one of them is the compressor unit. The compressor unit serves to increase the pressure needed for the next processing unit. In real conditions, several disturbances may occur in the process, affecting the stability of the system. Therefore, process control is needed for this system. Co-simulation is done by integrating Aspen plus dynamics for the nonlinear model of the compressor unit and MATLAB/Simulink for the control system model and mathematical calculation. Multivariable model predictive control (MMPC) is considered to control the system of the compressor unit. Three MPC parameters such as sampling time (T_s), prediction horizon (P), and control horizon (M) are set to 0.5, 50, and 30 seconds. The co-simulation gives various results. The highest overshoot is 19.6278 kPa in CV3 when SP on CV1 changed. The longest settling time occurred in CV4 when SP in CV4 was changed, 47.2729 seconds. The highest IAE is 14.8831, which occurred in CV4 when the SP of CV4 changed, and ISE is 200.7517 in CV3 when the SP of CV1 is altered.

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1. INTRODUCTION

Hydrogen is one of the substances that has many roles in various sector. In industry, hydrogen is widely used for petroleum refining, metal processing, fertilizer production, and food processing [1-3]. Hydrogen can also be used as an alternative energy source [4, 5]. Hydrogen is expected to be a contributor to energy transition and play an important role in the decarbonization of the global energy system [6-8]. In this research, the raw material of hydrogen production is based on biomass. This plant consists of raw material processing units, gasification units, char decomposer units, compression units, steam reforming units, char combustor units, cooling units, hydrogen sulphide (H_2S) removal units, and pressure swing absorber units [9-11]. The compressor unit serves to increase the outlet pressure of the gasification unit, in order to reach high pressure that required by H_2S removal unit.

The aim of this research is to co-simulate process control of nonlinear model of compressor unit. Nonlinear model. Nonlinear models are identical to the real situation compared to linear model. Simulation of process control on the compressor unit is carried out to maintain the optimal operating conditions of the plant. Additionally, this control aims to ensure the stability of plant operations in the event of external disturbances, thus preventing losses in terms of safety, financial, and damage to the equipment in the unit [12-14]. Model predictive control (MPC) with a multiple-input-multiple-output (MIMO) system or known as multivariable model predictive control (MMPC) is used as the controller of the system to maintain the stability the pressure of the compressor unit [15-19]. The simulations are done by using Aspen Plus Dynamics and MATLAB/Simulink software. The compressor unit is modelled on Aspen Plus Dynamics as nonlinear plant. The process control system is designed in

Simulink with the integration of Simulink and Aspen Plus Dynamics. The simulation and parameter computation are run by using MATLAB.

Several methods related to simulation of MPC to control a process are done by various researchers [20-23]. In 2020, Chinprasit and Panjapornpon proposed a co-simulation between Aspen Plus Dynamics and MATLAB/Simulink to simulate process control of vinyl chloride monomer (VCM) process [23]. The process consists of the ethylene dichloride (EDC) cracking, quenching, and purification sections, involving complex reactions, gas-liquid equilibrium, and multiple separators. The total of controlled variables (CV) is 11 and manipulated variables (MV) is 11, thus the MPC uses MIMO system. The linear plant model was built with state-space model. The results of the simulation are also compared with proportional-integral (PI) controller, which MPC provides better performance compared with the PI controller.

2. RESEARCH METHODS

2.1. Model Predictive Control

MPC offers several important advantages [24]: The process model captures the dynamic and static interactions between input, output, and disturbance variables; constraints on inputs and outputs are considered in a systematic manner; the control calculations can be coordinated with the calculation of optimum set points, and accurate model predictions can provide early warnings of potential problem. The structure of MPC is given in Figure 1.

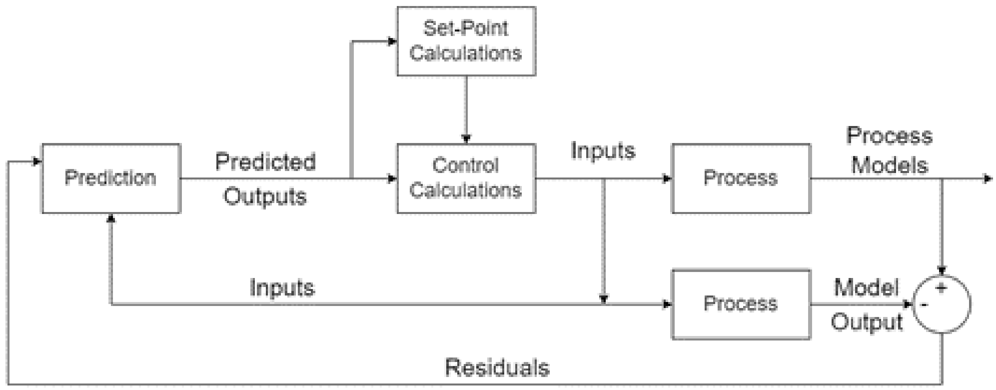


Figure 1. Diagram of model predictive control [24].

The purpose of the process model is used to predict the current values of the output variables. The interval between the actual and predicted outputs (residuals), serve as the feedback signal to a prediction block. The predictions are used in two types of MPC calculations that are performed at each sampling instant, those are set-point calculations and control calculations. The upper and lower limits of the calculation can be included in either type of calculation.The step-response model of a stable, single-input, single-output (SISO) process can be written as Equation (1):

$$y(k+1) = y_0 + \sum_{i=1}^{N-1} S_i \Delta u(k-i+1) + S_N u(k-N+1) \quad (1)$$

where, $y(k+1)$ is the output variable at the $(k+1)$-sampling instant, and $\Delta u(k-i+1)$ denotes the change in the manipulated input from one sampling instant to the next, $\Delta u(k-i+1) = u(k-i+1) - u(k-i)$. Both y and u are deviation variables. The model parameters are the N step-response coefficients, S_i to S_N. The initial value, y(0), is denoted by y_0. Three important parameters of MPC are sampling time (T_s), prediction horizon (P), and control horizon (M). Sampling time is the rate of the controller samples its inputs. The number of predictions P is referred as the prediction horizon and the number of control movements M is referred as the control horizon. For MIMO system, since the input and output consist of more than one variable, therefore, the equation can be written in matrix form in Equation (2):

$$\begin{bmatrix} y_1(s) \\ \cdot \\ \cdot \\ y_m(s) \end{bmatrix} = \begin{bmatrix} G_{p11}(s) & \dots\dots\dots. & G_{p1n}(s) \\ \vdots & \dots\dots\dots. & \vdots \\ G_{pm1}(s) & \dots\dots\dots. & G_{pmn}(s) \end{bmatrix} \begin{bmatrix} u_1(s) \\ \vdots \\ u_n(s) \end{bmatrix} \tag{2}$$

where, y_m is the m-th controlled variable, u_n is the n-th manipulated variable, and G_{pmn} is the process model for the m-th controlled variable and the n-th manipulated variable.

2.2. Linear Plant Creation

Although the plant simulation is done by using nonlinear model, the linear plant model is still needed for MPC. The model was generated with first order plus dead time (FOPDT) approach. The compressor unit model is shown in Figure 2, which consist of 4 unit. Each unit, consist of 1 compressor, 1 control valve, and 1 cooler.

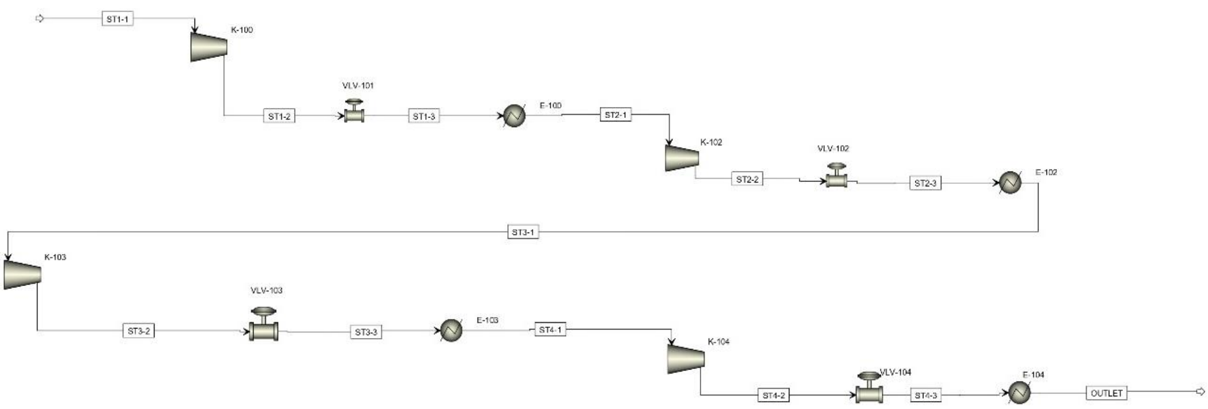


Figure 2. Compressor unit model.

Table 1. Controlled variable and manipulated variable of the simulation.

Notation	Controlled variable (kPa)	Notation	Manipulated variable (%)
CV1	Stream ST1-2 pressure	MV1	VLV-101 valve actuator
CV2	Stream ST2-2 pressure	MV2	VLV-102 valve actuator
CV3	Stream ST3-2 pressure	MV3	VLV-103 valve actuator
CV4	Stream ST4-2 pressure	MV4	VLV-104 valve actuator

The CV and MV are presented in Table 1. System identification was carried out by decreasing the valve opening by 20% to generate FOPDT model. The FOPDT model can be obtained with Equation (3) to (6) [12]:

$$G_p = \frac{K_P e^{-\theta s}}{\tau s+1} \tag{3}$$

$$K_p = \frac{\Delta y}{\Delta u} \tag{4}$$

$$\tau \ = 1{,}5\ (t_{63\%} - t_{28\%}) \tag{5}$$

$$\theta \ = t_{63\%} - \tau \tag{6}$$

where, G_p is the process model, K_p is process gain, Δy is change of CV, Δu is change of MV, τ is delay, θ is dead time, $t_{28\%}$ is the time when the process reaches 28% of the new value, and $t_{63\%}$ is the time when the process reaches 63% of the new value. The conversion plant signals from engineering units to a uniform dimensionless scale can simplify controller tuning significantly [25], thus, the unit of CV and MV from Aspen Plus Dynamics will be converted into dimensionless scale to simplify the calculation that executed in MATLAB/Simulink. System identification is also done by using dimensionless scale. The conversion is written with Equation (7) and (8):

$$y_{i-\%} = \frac{y_{i-nom} - y_{zeros}}{y_{span}} \times 100\% \tag{7}$$

$$y_{span} = 2 \times y_{0-nom} \tag{8}$$

where, $y_{i-\%}$ is dimensionless CV at i-iteration, y_{i-nom} is nominal value CV at i-iteration, y_{zeros} is zero value, y_{0-nom} is the default nominal value of y, and y_{span} is y_{0-nom} multiplied by two. Therefore, when there is no change of set-point (SP), the default value of $y_{i-\%}$ is 50%. Equation (4) and (5) are also applied for the conversion of dimensionless MV but for MV the symbol is notated by u and there is no set-point. Table 2 present the result of system identification parameters using FOPDT approach.The graph of model validation result will be presented in form of dimensionless amplitude, so the final CV value $y_{f-\%}$ is subtracted by 50%.

$$Amplitude = y_{f-\%} - 50\% \tag{9}$$

2.3. Control System Model

The control system model is built in Simulink and illustrated in Figure 3, which consist of MPC as the controller, Set-point block to define the SP of the process, unit conversion block to convert engineering and percentage unit, and AMSimulation to integrate Simulink with Aspen Plus Dynamics.

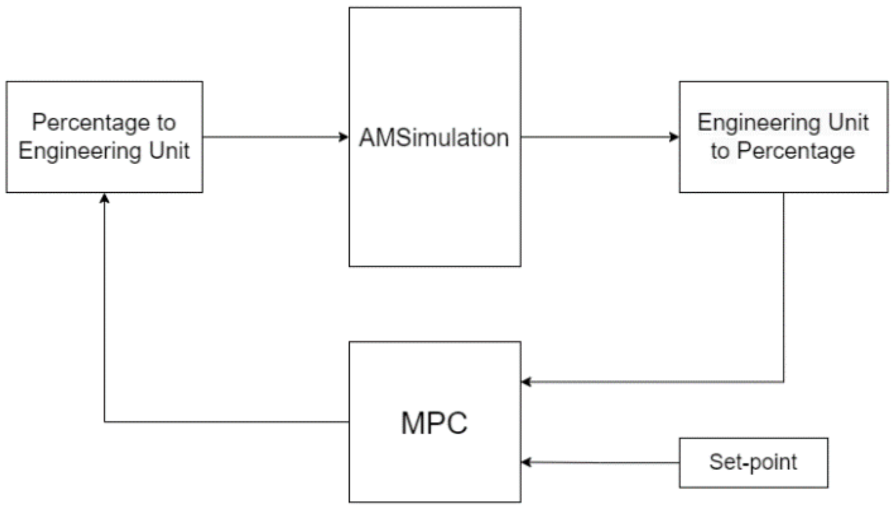


Figure 3. Diagram of compressor unit control system.

Later, the co-simulation will be executed in MATLAB by calling the Simulink file due to the MPC must be built with various parameters.

2.4. Simulation Parameters

In this research, several parameters are used to review the performance of MPC. Integral of the absolute error (IAE) and integral squared error (ISE) are applied to calculate the error. Other parameters such as overshoot and settling time are also applied to inspect the exceeding CV value and the time needed for the process to reach the new SP. The SP is increased 10 kPa from the initial pressure in each unit compressor,which given in Table 3. Subsequently, the final SP is converted into dimensionless scale. The test is performed one-by-one by changing one of the SP on the compressor unit. Hence, the tests are done 4 times.

Table 2. SP changes parameters.

No.	Initial SP (y_{0-nom}) (kPa)	Final SP (kPa)	Initial SP (%)	Final SP (%)
1	333.2993	340.2993	50	51.050
2	719.2944	726.2944	50	50.487
3	1601.9894	1608.9894	50	50.218
4	3174.0009	3181.0009	50	50.110

Three important MPC parameters such as T_s, P, and M are set to 0.5, 50, and 30 second respectively. The error IAE and ISE can be calculated with Equation (10) and (11) [12]:

$$IAE = \int |SP(t) - CV(t)|\, dt \tag{10}$$

$$IAE = \int \left(SP(t) - CV(t)\right)^2 dt \tag{11}$$

3. RESULTS AND DISCUSSIONS

3.1. Model Creation and Model Validation Results

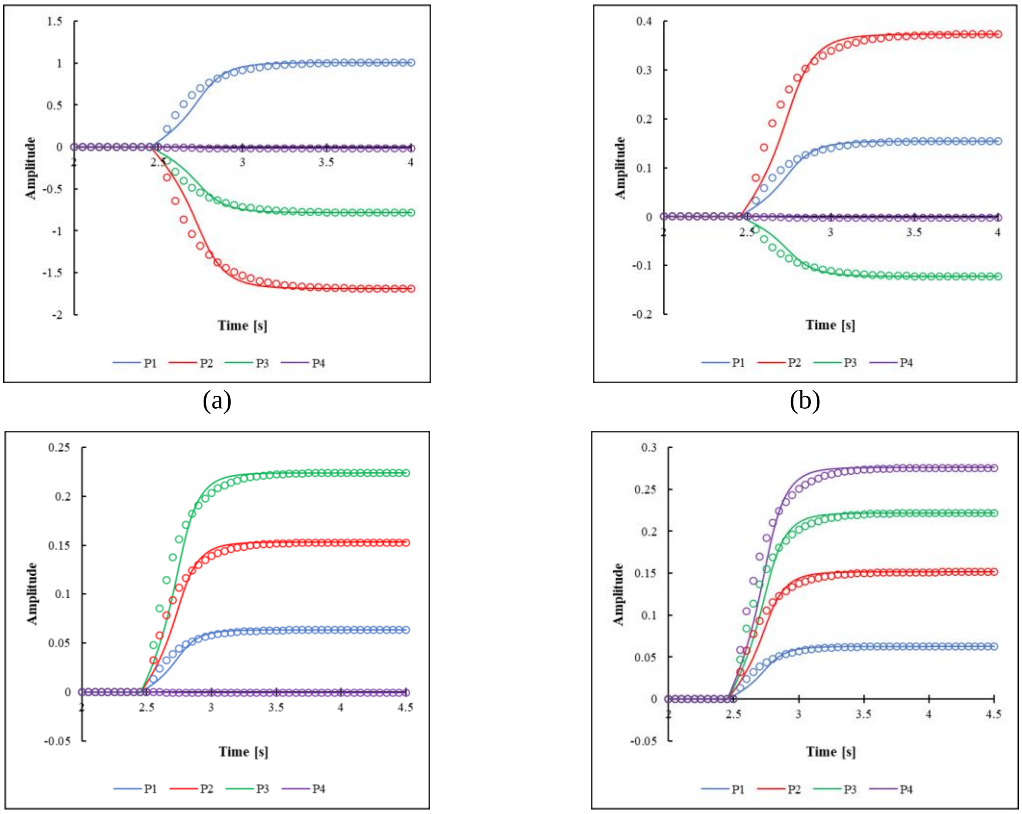
To obtain the data for FOPDT modelparameters calculation, the system identification is run from Simulink, so the signal transmitted from the aspen plus dynamics can be logged. Next, the logged data is processed in Microsoft Excel, generating FOPDT model parameters, such as Kp, τ, and θ. Those parameters are given in Table 3.

Table 3. FOPDT model parameters.

Model		Kp	τ (s)	θ (s)
G1	G1.1	-0.0505	0.2099	0.1071
	G1.2	-0.0078	0.2096	0.1097
	G1.3	-0.0032	0.2095	0.1103
	G1.4	-0.0031	0.2095	0.1101
G2	G2.1	0.0846	0.2103	0.1041
	G2.2	-0.0187	0.2095	0.1098
	G2.3	-0.0076	0.2095	0.1103
	G2.4	-0.0076	0.2095	0.1101
G3	G3.1	0.0393	0.2102	0.1050
	G3.2	0.0061	0.2096	0.1094
	G3.3	-0.0112	0.2094	0.1104
	G3.4	-0.0111	0.2095	0.1102
G4	G4.1	0.0008	0.1861	0.1268
	G4.2	0.0001	0.2097	0.1090
	G4.3	0.0001	0.2095	0.1100
	G4.4	-0.0138	0.2095	0.1102

After FOPDT model was obtained, the model validation nis also carried out to decide whether the model can be applied to MPC or not.Methods of the model validation tests are same as the system identification, by changing MV in each unit by -20% from its initial value. The linear model is executed in MATLAB and the nonlinear model is executed in Aspen Plus Dynamics. The model validation results are illustrated in form of graph in Figure 4.

Based on Figure 4, despite the difference when the pressure is increasing or decreasing,the linear and nonlinear model are almost identical and give the samebehavior and results at stable condition. For example, when changing MV1, CV1 of linear model and nonlinear model are increasing, while CV2, CV3, and CV4 are decreasing. Hence, the FOPDT will be applied to MPC.

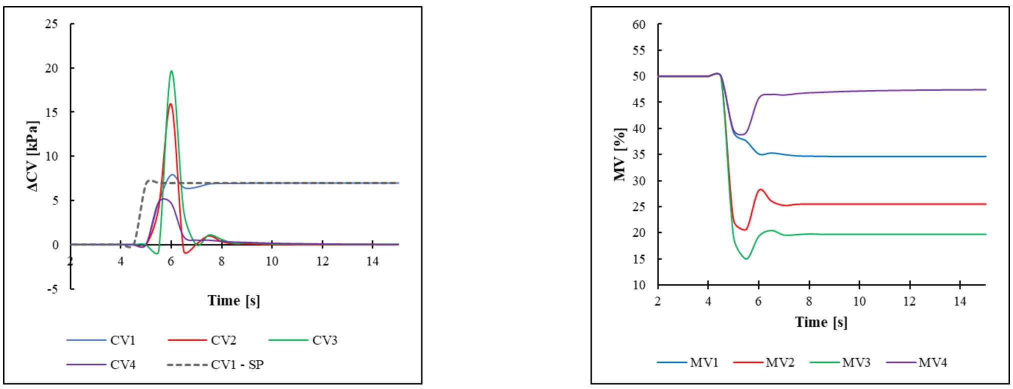


(a) (b) (c) (d)

Figure 4. Model validation test results by changing (where, linear: dot and nonlinear: line): (a) MV1; (b) MV2; (c) MV3; and (d) MV4 by decreasing 20% of its initial value.

3.2. Process Control Simulation

The tests are carried out by changing the SP at the compressor output in each compressor unit, giving various results in each test. Each test, the simulation start on 0 second and ends on 50 seconds with 0.5 seconds interval. Overall, when the valve opening is increased, the stream pressure is decreased, and the mass or molar flow are increased. For SP change test on CV1, the results is given in Figure 5 and Table 4. Based on this test, the controller isable to stabilize the stream into its new condition. When the CV1 pressure is increased, CV2 and the other next streams pressureare decreased.Thus, CV2 and the other next streams pressure need to beincreased. For that reason, final MV of all streams are decreased. The highest overshoot occurred in CV3 and also has the highest error.



(a) (d)

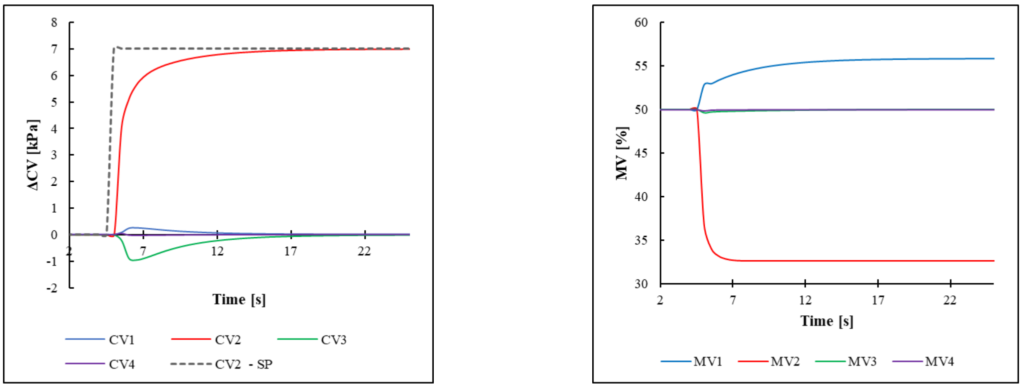
Figure 5. Simulation results of SP change on CV1: (a) CV and (b) MV response.

Table 4. Performance of MMPC with SP change on CV1.

Parameter	CV1	CV2	CV3	CV4
Overshoot (kPa)	0.9157	15.8652	19.6278	4.7636
Settling time (s)	8.6975	-	-	-
IAE	5.7078	11.3005	13.3455	6.6364
ISE	27.6383	135.1228	200.7517	23.0653

For SP change test on CV1, the results is given in Figure 5 and Table 4. When the SP on CV2 increased, CV1 is increased, but CV3 and CV4 are decreased. So, the controller responses by increasing MV1 and decreasing MV2, MV3, and MV4. In this case, beside MV2, MV1 is the most affected by the change of SP on CV1. The test gives different controller performance results compared to the previous test. Overshoot only occurred on CV1 and slightly on CV2. On CV3 and CV4, the flow pressures undershoot occurs, due to the decreased pressure before its stable condition. The undershoot or overshoot on this test also has lower deviation compared to SP change on CV1 test. As on the settling time, the stream pressure reaches its new SP faster than the previous test. Overall, the error on this test has smaller value than the previous test. For SP change test on CV2, the results is given in Figure 5 and Table 4.

The test results of SP changes test on CV3 are presented in Figure 7 and Table 6. This test result is giving identical pattern to SP change on CV1 test, which no undershoot occurred. When CV3 is increased, CV1 and CV2 is increased, however CV4 is decreased. So, the controller responses by increasing MV1 and MV2, and decreasing MV3 and MV4, but MV1 and MV4 return to its initial condition (50%). Overshoot in this test is significantly smaller than on SP changes on CV1 test, so does the error, but the errors are identical to SP changes on CV2 test.



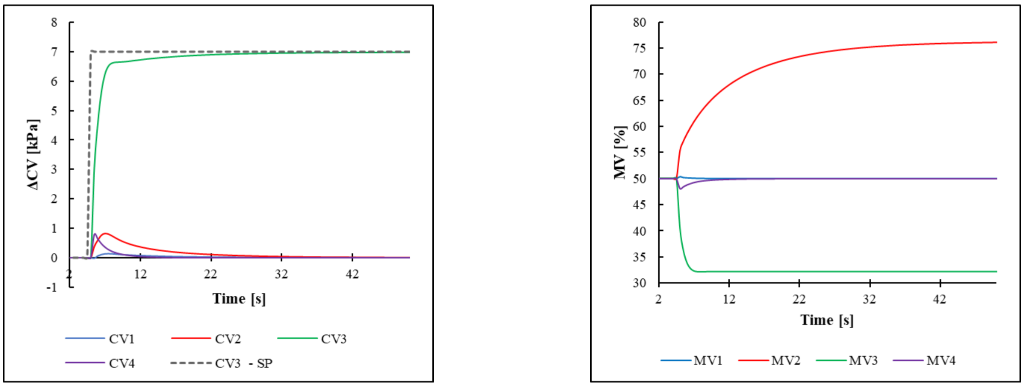
(a) (d)

Figure 6. Simulation results of SP change on CV2: (a) CV and (b) MV response.

Table 5. Performance of MMPC with SP change on CV2.

Parameter	CV1	CV2	CV3	CV4
Overshoot (kPa)	0.2624	-	-0.9452	-0.0300
Settling time (s)	-	17.7735	-	-
IAE	1.2489	10.0740	4.3984	0.1712
ISE	0.1970	33.3095	2.5061	0.0033

The SP change on CV4 test results are presented in Figure 8 and Table 7. This test also does not give undershoot. The SP change on CV4 is heavily affecting MV3, as can be seen on the figure, the MV3 is still increasing while MV4 already in its stable condition. Meanwhile, MV1 and MV2 are almost no affected. This test has the longest settling time compared to other test.



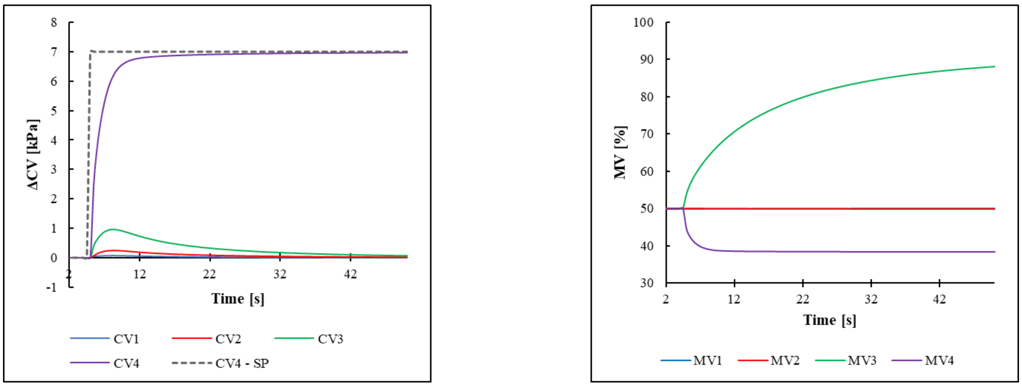
(a) (d)

Figure 7. Simulation results of SP change on CV3: (a) CV and (b) MV response.

Table 6. Performance of MMPC with SP change on CV3.

Parameter	CV1	CV2	CV3	CV4
Overshoot (kPa)	0.1365	0.8233	-	0.7901
Settling time (s)	-	-	36.9396	-
IAE	1.2911	7.0213	12.4075	1.8515
ISE	0.0929	3.0126	37.5263	0.7935

Overall, stream pressure change in compressor unit 1 is affecting the other streams more than change in unit 2, 3, and 4, therefore giving the highest error. The stream pressure change in compressor unit 1 also gives the highest overshoot than changes in other unit.



(a) (d)

Figure 8. Simulation results of SP change on CV4: (a) CV and (b) MV response.

Table 7. Performance of MMPC with SP change on CV4.

Parameter	CV1	CV2	CV3	CV4
Overshoot (kPa)	0.0729	0.2456	0.9590	-
Settling time (s)	-	-	-	47.2729
IAE	1.1381	3.7665	14.8279	14.8831
ISE	0.0477	0.5167	8.0375	46.0959

4. CONCLUSION

MPC is able tocontrol the process incompressor unit stream, giving various results. The highest overshoot is 19.6278 kPa occurred in CV3 when the SP on CV1 changes and undershoot only occurred when SP on CV2 changes. Changes of SP in CV4 has longest settling time, 47.2729 seconds. The highest IAE is 14.8831, shown on CV4 when SP of CV4 changed and ISE is 200.7517, occurred at CV3 when SP of CV1 is changed.

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REFERENCES

[1] Manna, J., Jha, P., Sarkhel, R., Banerjee, C., Tripathi, A. K., & Nouni, M. R. (2021). Opportunities for green hydrogen production in petroleum refining and ammonia synthesis industries in India. *International Journal of Hydrogen Energy*, **46**(77), 38212–38231.

[2] Xu, X., Zhou, Q., & Yu, D. (2022). The future of hydrogen energy: Bio-hydrogen production technology. *International Journal of Hydrogen Energy*, **47**(79), 33677–33698.

[3] Qazi, U. Y. (2022). Future of hydrogen as an alternative fuel for next-generation industrial applications; challenges and expected opportunities. *Energies*, **15**(13), 4741.

[4] EIA, U. (2022). *Hydrogen explained*. Production of Hydrogen.

[5] Hosseini, S. E. & Wahid, M. A. (2020). Hydrogen from solar energy, a clean energy carrier from a sustainable source of energy. *International Journal of Energy Research*, **44**(6), 4110–4131.

[6] Pribadi, A. (2022). *Hidrogen didorong jadi kontributor transisi energi Indonesia*. Kementerian Energi dan Sumber Daya Mineral.

[7] Zhang, Y., Davis, D., & Brear, M. J. (2022). The role of hydrogen in decarbonizing a coupled energy system. *Journal of Cleaner Production*, **346**, 131082.

[8] Quarton, C. J., Tlili, O., Welder, L., Mansilla, C., Blanco, H., Heinrichs, H., Leaver, J., Samsatli, N. J., Lucchese, P., Robinius, M., & Samsatli, S. (2020). The curious case of the conflicting roles of hydrogen in global energy scenarios. *Sustainable Energy & Fuels*, **4**(1), 80–95.

[9] Budianta, I. A., Abqari, F., & Cicilia, A. (2011). *Pabrik bio-hidrogen dari biomasa*. Department of Chemical Engineering of Universitas Indonesia, Universitas Indonesia.

[10] Shamsi, M., Obaid, A. A., Farokhi, S., & Bayat, A. (2022). A novel process simulation model for hydrogen production via reforming of biomass gasification tar. *International Journal of Hydrogen Energy*, **47**(2), 772–781.

[11] Dumbrava, I. D. & Cormos, C. C. (2021). Techno-economical evaluations of decarbonized hydrogen production based on direct biogas conversion using thermo-chemical looping cycles. *International Journal of Hydrogen Energy*, **46**(45), 23149–23163.

[12] Marlin, T. E. (2000). *Process control: Designing processes and control systems for dynamic performance*. McGraw-Hill, Singapore.

[13] Ilyushin, P., Filippov, S., Kulikov, A., Suslov, K., & Karamov, D. (2022). Intelligent control of the energy storage system for reliable operation of gas-fired reciprocating engine plants in systems of power supply to industrial facilities. *Energies*, **15**(17), 6333.

[14] Ilyushin, P., Filippov, S., Kulikov, A., Suslov, K., & Karamov, D. (2022). Specific features of operation of distributed generation facilities based on gas reciprocating units in internal power systems of industrial entities. *Machines*, **10**(8), 693.

[15] Du, J., Chen, J., Li, J., & Johansen, T. A. (2021). Multiple Model Predictive Control for Nonlinear Systems Based on Self-Balanced Multimodel Decomposition. *Industrial and Engineering Chemistry Research*, **61**(1), 487–501.

[16] Zhang, J., Chin, K. S., & Ławryńczuk, M. (2017). Multilinear model decomposition and predictive control of mimo two-block cascade systems. *Industrial and Engineering Chemistry Research*, **56**(47), 14101–14114.

[17] Schwenzer, M., Ay, M., Bergs, T., & Abel, D. (2021). Review on model predictive control: An engineering perspective. *The International Journal of Advanced Manufacturing Technology*, **117**(5), 1327–1349.

[18] Wahid, A., Meizvira, F., & Wiranoto, Y. (2018). Application of multivariable model predictive control to overcome the intervariable interaction in CO2 removal process. *E3S Web of Conferences*, **67**, 03049.

[19] Akinola, T. E., Oko, E., Wu, X., Ma, K., & Wang, M. (2020). Nonlinear model predictive control (NMPC) of the solvent-based post-combustion CO2 capture process. *Energy*, **213**, 118840.

[20] Eggins, J. (2015). *Integration of MATLAB and LabVIEW with Aspen Plus Dynamics*. Research Thesis, Instrumentation and Control, Electrical Power, Mudorch University.

[21] Tuan, T. T., Tufa, L. D., Mutalib, M. I. A., & Abdallah, A. F. M. (2016). Control of depropanizer in dynamic Hysys simulation using MPC in Matlab-Simulink. *Procedia Engineering*, **148**, 1104–1111.

[22] Wahid, A. & Taqwallah, H. M. H. (2018, March). Model predictive control based on system re-identification (MPC-SRI) to control bio-H2 production from biomass. *IOP Conference Series: Materials Science and Engineering*, **316**(1), 012061.

[23] Chinprasit, J. & Panjapornpon, C. (2020). Model predictive control of vinyl chloride monomer process by Aspen Plus Dynamics and MATLAB/Simulink co-simulation approach. *IOP Conference Series: Materials Science and Engineering*, **778**(1), 012080.

[24] Conner, J. S. & Seborg, D. E. (2005). Assessing the need for process re-identification. *Industrial and Engineering Chemistry Research*, **44**(8), 2767–2775.

[25] Anon, A. (2023). *Design and Cosimulate Control of High-Fidelity Distillation Tower with Aspen Plus Dynamics (Mathworks).* Aspen Plus Dynamics (Mathworks).